

Membrane Distillation and Applications for Water Purification in Thermal Cogeneration—A Pre-study

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Membrandestillering och tillämpningar för vattenrening i kraftvärmeverk – Förstudie

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Abstract

The objective of the present investigation is to explore the feasibility of membrane distillation (MD) as a complimentary or replacement technology for water purification processes in thermal cogeneration plants. This report contains background information on MD, select experimental results, and a case study of industrial applications. Results show that current MD technology features similar energy consumption levels and higher specific costs as compared to reverse osmosis. Prospects for improvement in the near future are judged to be positive regarding these and other aspects.

Sammanfattning

Kostnadseffektiv, pålitlig och energisnål vattenreningsteknik är en viktig del i moderna kraftvärmeverk. Avsaltat vatten behövs som spädvatten i fjärrvärmenät samt som processvatten i pannor och turbinanläggningar. Dessutom har det blivit aktuellt med rening och återvinning av rökgaskondensat. Idag finns det flera lämpliga tekniker såsom omvänd osmos (RO) och elektroavjonisering (EDI). Membrandestillering (MD) är en ny lovande teknik i sammanhanget. Denna teknik utnyttjar partialtryckdifferenser för att rena vatten med hjälp av hydrofoba membran. Processen kan drivas av fjärrvärme eller lågtemperaturånga och är därmed attraktiv i kombination med kraftvärmeproduktion. Denna förstudie fokuseras på membrandestillering som en ny vattenreningsteknik i kraftvärmeverk. Resultaten kommer att ligga till grund för eventuellt fortsatt forskning som exempelvis kan omfatta pilotstudier. Till målgrupperna hör miljötekniker och driftoperatörer med intresse för ny teknik.

Förstudien omfattar en litteraturstudie, teori gällande mass- och värmetransport samt uppskalning av experimentella resultat. Data från en testanläggning som ägs av Xzero AB och som finns på KTH har använts. Den experimentella renvattenproduktionen låg under den teoretiska maximala gränsen vilket visar att det finns potential för förbättringar vad gäller MD-modulens design. En fallstudie med ett system som producerar 10 m³ renvatten/h valdes för att belysa kommersiella aspekter. Studien visar att MD är en lovande alternativ till RO i befintliga eller nya anläggningar. De bästa alternativen är då värme från fjärrvärmeframledningen eller lågtemperaturånga (2-3 bar, 200°C) utnyttjas. Det specifika energibehovet för MD är enligt följande: 4,0-5,0 kWh värme/m³ producerat renvatten och 1,5-4,0 kWh el/m³ renvatten. Även om det totala energibehovet är större jämfört med RO, kan förbättringar i processen reducera eller eliminera denna nackdel. Elförbrukningen är huvudsakligen kopplad till MD-anläggningens höga interna vattenrecirkulering, som kan minimeras i framtiden. Därmed visar MD tillfredställande energiprestanda i förhållande till dagens teknik. De specifika kostnaderna ligger kring 10-14 SEK/m³ renvatten för de mest sannolika konfigurationerna. MD:s nuvarande ställning är därmed något svag ekonomiskt jämfört med RO, dock bör man väga in respektive tekniks utvecklingsgrad i en mer rättvis jämförelse.

Förstudiens lovande resultat motiverar fortsatt forskning. Pilotförsök rekommenderas för att ytterligare utvärdera teknikens potential, framförallt när det gäller vattenkvalitet och ekonomi. Grundläggande undersökningar med mass- och värmetransport i verkliga moduler föreslås, eftersom det finns utrymme för ökat utbyte och/eller minskad membranåta. Med fortsatt teknikutveckling och -optimering bör MD kunna utvecklas till en mer konkurrenskraftig teknik.

Nyckelord: membrandestillering; vattenteknik; spädvatten; rökgaskondensat; spillvärme.

Summary

Cost-effective, reliable, and energy efficient water treatment systems are an integral part of modern cogeneration facilities. Demineralized water is required for make-up water in district heating networks and in boilers. In addition, increasing attention has been paid to the treatment of flue gas condensate for possible recycling. A number of membrane technologies like reverse osmosis (RO) and electrodeionization (EDI) have been developed for the above applications. Besides these methods, membrane distillation (MD) is promising technology in this context. MD utilizes differences in vapor pressure to purify water via a hydrophobic membrane. The process can utilize district heat supply temperatures or low-grade steam, thus making it attractive for cogeneration applications. This investigation consists of a pre-study to evaluate the viability of membrane distillation as a new water treatment technology in cogeneration plants. Results obtained from the study will be used as an input to follow-on research, which may include the construction of a pilot plant. Target groups for this study include environmental engineers with particular interest in emerging water purification technologies.

Specific elements of this work include a literature survey, theoretical considerations of heat and mass transfer, and scale-up of experimental results. Data obtained from the test facility owned by Xzero AB and located at KTH was employed for this purpose. Actual water production was found to be lower than the theoretical maximum, illustrating the potential for improvements in MD module design. A case study considering a 10 m³ pure water/hr system is explored to shed light on commercial-scale aspects. Results show that MD is a promising alternative to RO in existing or new treatment facilities. The most favorable results were obtained for alternatives where either the district heat supply line or low-grade steam (2-3 bar, 200°C) are available. Specific energy consumption ranges are as follows: 4.0-5.0 kWh/m³ thermal; and 1.5-4.0 kWh/m³ electrical. The relatively high electricity consumption is linked primarily to high recirculation rates versus relative low water production in batch mode. Although the combined energy consumption is higher than RO, future process improvements can be employed to offset or eliminate this disadvantage. MD thus demonstrates satisfactory energy performance compared to existing technologies. Specific costs lie in the range of 10-14 SEK/m³ for the most likely MD system scenarios. These results indicate that MD is presently more expensive than RO, although this comparison should be weighed against the level of development for each respective technology.

These promising results suggest that follow-on research is justified. Pilot plant trials are recommended in order to fully assess the potential of this technology, especially with regards to water quality issues and economic viability. Fundamental investigations considering heat and mass transfer in real membrane modules are also suggested, since there is potential for enhancing pure water production and minimizing membrane area. With continued research and development the possibility of commercializing MD technology in the near-term will become even more likely.

Keywords: membrane distillation; water treatment; make-up water; flue gas condensate; low-grade heat utilization.

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1. Introduction

Thermal cogeneration plants require purified or treated water for a number of processes. The selection of water treatment technology is dependent upon the final water quality along with the volume of water to be treated. A brief overview of current practices is listed below:

Feedwater/make-up water treatment in district heating networks: Water quality should be relatively high in order to avoid corrosion in piping and heat exchangers. Treatment steps include filtration, demineralization (reverse osmosis, ion exchange), deaeration, and pH adjustment via addition of NaOH [1].

Make-up water purification, steam cycle: Metallurgical constraints place a relatively high demand on water quality. A number of processes can be used depending on the particular requirements, including chemical treatment/lime softening, filtration, carbon adsorption, reverse osmosis (RO), and ion exchange resin polishing [2]. The amount of water to be treated depends upon the size of the plant; for example, the Örebro Cogeneration Facility with a fuel capacity of about 500 MW requires 45 000 m³/yr for its boilers [3].

Condensate treatment, flue gas condensation: Wastewater treatment is more complicated and is highly dependent upon the fuel characteristics. The possible presence of heavy metal compounds requires precipitation/flocculation and clarification. Filtration and chemical treatment is required prior to discharge to receiving water [4].

Although the above methods are well established and generally effective, there is still room for improvement regarding availability, economy, and enhanced water purity. Technologies of particular interest include advanced membranes in reverse osmosis (RO) and electrodeionization (EDI) [5].

Beyond these methods, membrane distillation is a promising alternative or complementary technology in water treatment systems of thermal cogeneration facilities. Membrane distillation (MD) is a hybrid process that employs specially designed hydrophobic membranes for the unit operation, and mass transport is driven by differences in vapor pressure [6]. Temperature levels are such that low-grade heat sources (e.g. district heat) may be used to supply the required energy to the process. Unlike other membrane processes, MD does not require high pressures and is not limited by the osmotic pressure. In addition MD differs from other distillation processes in that vacuum pressures and/or high temperatures are not necessarily required.

In short, an MD unit consists of heating and cooling loops connected to a module containing the membranes (which are usually arranged in a series of vertical channels). Raw water is heated and supplied to the hot side, while purified water is collected on the cold side after evaporation through the membrane. Figure 1 shows a schematic diagram of the MD process at the membrane level.

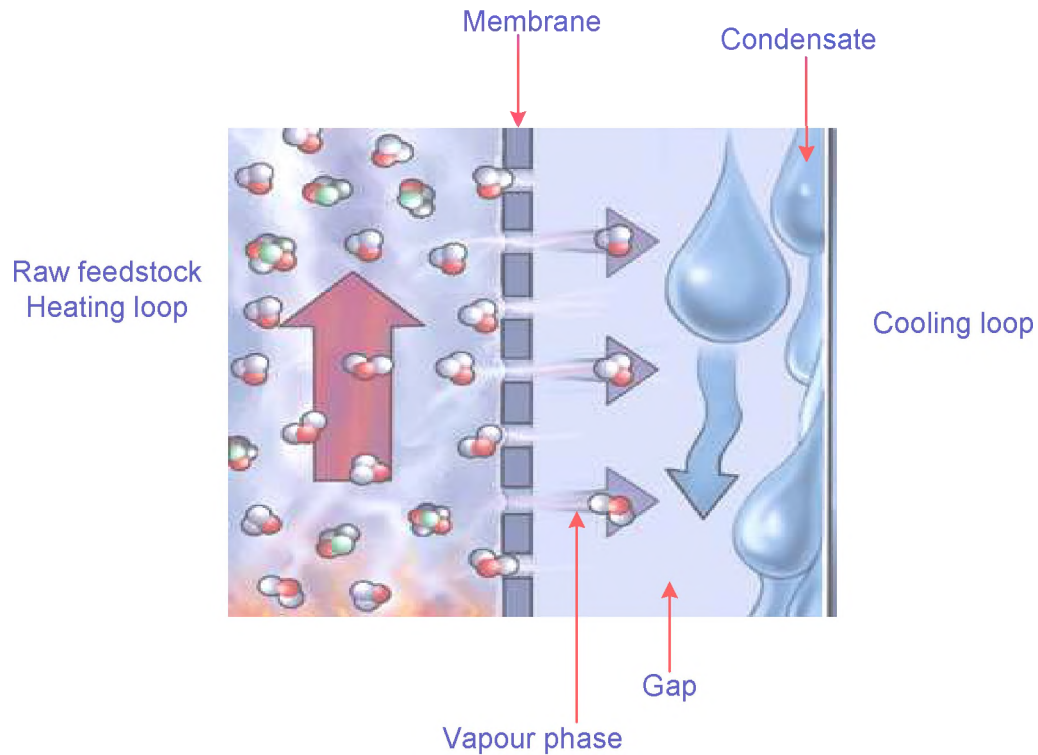


Figure 1. Schematic of membrane distillation process, Xzero AB

Water is purified via the following steps [7]:

- Heat is transported from the bulk fluid to the membrane surface, and nonvolatile components diffuse from the membrane surface to the bulk
- Water in the hot containment section evaporates
- Water vapor diffuses through the membrane
- Water vapor diffuses through the gap, from the membrane surface to the condensate wall
- Water vapor condenses on the condensate wall
- Heat is transported through the condensate wall to the coolant

Advantages of MD:

- 100% (theoretical) rejection of ions, macromolecules, colloids, cells, and other non-volatiles
- Lower operating temperatures than conventional distillation
- Lower operating pressure than conventional pressure-driven membrane separation processes
- Low sensitivity to variations in process variables (e.g. pH and salts [8])
- Good to excellent mechanical properties and chemical resistance
- Potentially lower capital costs as compared to RO

Disadvantages of MD:

- High energy intensity (although energy, i.e. heat, is usually low grade)
- Low yield in non-batch mode; high recirculation rates in batch mode
- Sensitive to surfactants
- Volatiles cannot be completely separated (degassing or other methods required, however, volatiles may be captured in the air space which is then vented off)
- Not yet commercially available for power plant applications

1.1 Description of the research area

Within the Värmeforsk framework, water treatment issues have been studied primarily in the context of flue gas condensation [e.g. 10-17]. Here, RO has been the primary focus [10-11]. One ongoing Värmeforsk investigation, *Avancerad rening av rökgaskondensat* [16] is exploring aspects related to clogging or deterioration of membranes in RO systems. New strategies for pre-treatment are of particular interest, and tests will be conducted in pilot scale equipment. Another Värmeforsk investigation, *Ny vattenreningsteknik för energianläggningar – Det europeiska vattendirektivets praktiska konsekvenser för värmeverken* [17] considers new water treatment methods in the context of modified European regulations. Research on MD technologies appears to be a good complement to these studies.

1.2 The purpose of the research assignment and its role within the research area

This investigation is divided into two parts. Part 1 (the present work) consists of a pre-study to evaluate the viability of membrane distillation as a new water treatment technology in thermal cogeneration plants. Specific goals include the following:

- Literature survey of existing and future water treatment technologies in the present context
- Identification of specific applications for cogeneration facilities
- Mass and energy balance analysis
- Dimensioning of membrane distillation equipment
- Economic analysis of installation and operation and maintenance costs

Part 2 (to be proposed) will focus on experimental investigations (pilot plant trials) where MD's performance and applicability for make-up water production is evaluated. Parametric variations for typical raw water and/or flue gas condensate will be considered in detail. Water quality issues are anticipated to comprise a significant element of follow-on work.

2. Membrane distillation water purification technology

2.1 Background

Membrane distillation (MD) is a novel water purification process being investigated worldwide as a low cost, energy saving alternative to conventional separation processes, such as distillation. The benefits of MD compared to other more popular separation processes are the following: (1) 100% (theoretical) rejection of ions, macromolecules, colloids, cells, and other non-volatiles, (2) lower operating temperatures than conventional distillation, (3) lower operating pressure than conventional pressure-driven membrane separation processes, (4) reduced chemical interaction between membrane mechanical property requirements, and (5) reduced vapor spaces compared to conventional distillation processes [19]. This thermally driven process employs a hydrophobic microporous membrane to support a vapor-liquid interface. If a temperature difference is maintained across the membrane, a vapor pressure difference occurs. As a result, liquid (usually water) evaporates at the hot interface, crosses the membrane in the vapor phase and condenses at the cold side, giving rise to a net transmembrane water flux.

A variety of methods may be employed to impose this vapor pressure difference, and in general there are four kinds of MD system configurations (Figure 2) e. g. direct contact membrane distillation (DCMD), air gap membrane distillation (AGMD), sweep gas membrane distillation (SGMD) and vacuum membrane distillation (VMD).

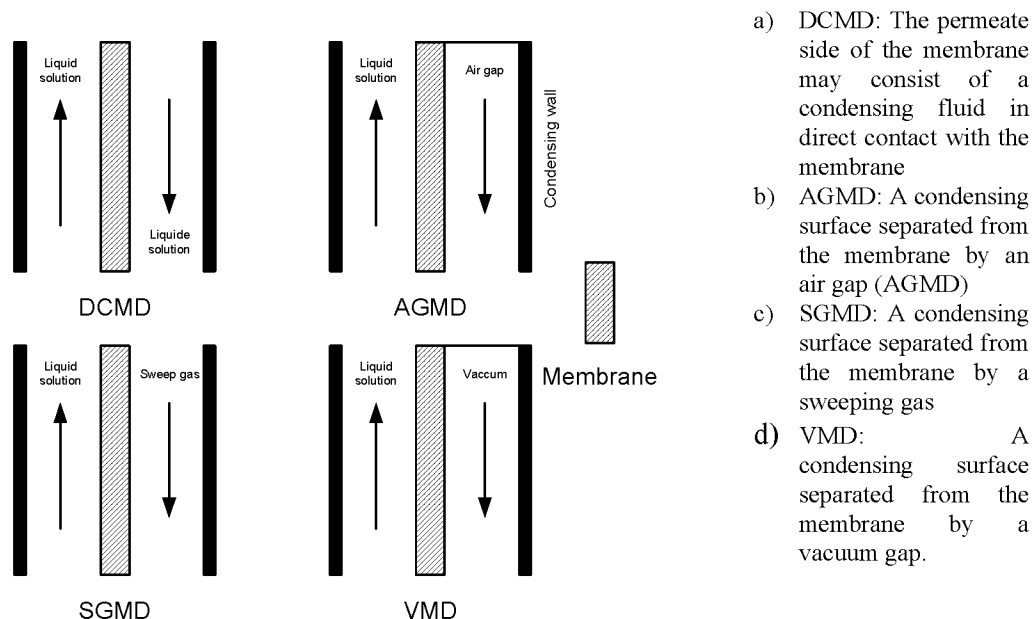


Figure 2. Common configurations of membrane distillation [19]

DCMD and AGMD are the most popular methods, but one of the problems with DCMD is the relatively low efficiency of heat utilization. A large portion of the heat supplied to the feed solution is lost by conduction through the membrane. Another reason why DCMD is unsuitable for water treatment applications is the need to use pure water for cooling.

In the present work, air gap membrane distillation (AGMD) alone is investigated. The principle advantage of AGMD against other configurations of membrane distillation arises from the possibility of condensing the permeate vapors on a cold surface rather than directly in a cold liquid. In this configuration, the mass transfer steps involve movement within the liquid feed toward the membrane surface, evaporation at the membrane interface and transport of the vapor through the membrane pores and air gap prior to condensation [20].

The most important advantage of the MD process is that it does not need to operate at high temperature as compared to the traditional distillation process, nor are high pressures required as compared to reverse osmosis. Thus low-grade heat (theoretically temperatures in the range from 30 to 70°C) can be employed to drive the MD process. Such temperatures are available from waste heat streams in the power plant or can be obtained from small-scale cogeneration units.

2.2 Numerical Model describing AGMD process

The mathematical model presented by Jönsson et al. [21] is considered for the present study. This model considers one-dimensional, steady state heat and mass transport through a hydrophobic membrane. A summary of this model is given below:

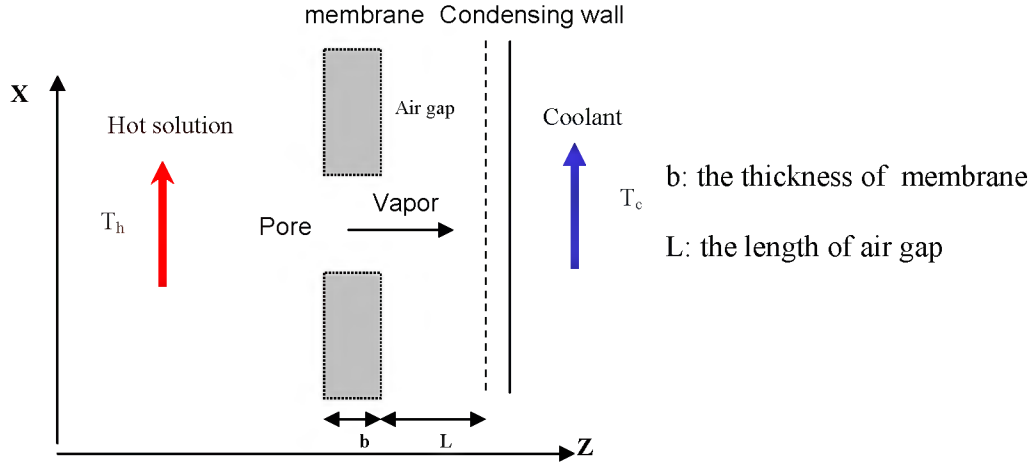


Figure 3. Schematic representation of the membrane distillation module

As shown in Figure 3, the hot solution flows in the x direction and some mass is lost in the z direction due to evaporation. At steady state the molar flux, N , of a vapor diffusing through a stagnant air gap is described as [7]:

$$N = -\frac{CD}{1-x} \left(\frac{dx}{dz} \right) \quad (1)$$

where C is the molar concentration and D is the diffusion coefficient for the water vapor-air mixture. Although there is a slight effect on the molar flux, the simultaneous mass transfer most directly affects the rate of heat transfer. In this case, the sensible heat, E , includes two terms encompassing conductive and diffusive contributions:

$$E = -k \frac{dT}{dz} + NC_p (T - T_c) \quad (2)$$

where k is the effective thermal conductivity, C_p is the heat capacity and T_c is the absolute temperature of coolant. Differential equations (2) and (3) are solved in the region of membrane and the air gap. The factors $C * D$ and air conductivity k_{air} are obtained according to two empirically derived relations valid for water vapor-air mixtures at around 40°C:

$$C * D = 6.3 * 10^{-5} \sqrt{T} \quad (3)$$

$$k_{air} = 1.5 * 10^{-5} \sqrt{T} \quad (4)$$

The effective thermal conductivity coefficient k for the membrane region is assumed to be obtained from standard porous media relations [22], i.e.

$$k = k_{air} * \phi + k_{membrane} (1 - \phi) \quad (5)$$

where ϕ and $k_{membrane}$ are the membrane porosity and thermal conductivity. It is possible to obtain explicit expressions for molar and energy fluxes when the above relations are substituted into Equations (4) and (5). Although these expressions have no exact analytical solutions, there are two empirical equations that yield sufficiently accurate values, i.e.:

$$N = 6.3 * 10^{-5} * \frac{1}{b / (\phi * \sqrt{T_h}) + L / \sqrt{T_c}} * \ln\left(\frac{1 - V_c}{1 - V_h}\right) \quad (6)$$

and

$$E = \frac{1.5 * 10^{-3} * (T_h - T_c)}{(b / \gamma * \phi * \sqrt{T_h}) + L / \sqrt{T_c}} * 1 + 1.41 * \ln\left(\frac{1 - V_c}{1 - V_h}\right) * \frac{b / (\gamma * \phi * \sqrt{T_h})}{(\frac{b}{\phi \sqrt{T_h}}) + (L / \sqrt{T_c})} \quad (7)$$

where

$$\gamma = \frac{k}{\phi k_{air}} \quad (8)$$

and k and k_{air} should be evaluated at T_h . In the above, V_c is mole fraction of water vapor at the condensate surface and V_h is the mole fraction of water vapor at the evaporation surface. The mass flux equation is obtained from the molar flux equation:

$$Q = 4.1 * 10^{-3} * \frac{1}{b / (\phi * \sqrt{T_h}) + L / \sqrt{T_c}} * \ln\left(\frac{1 - V_c}{1 - V_h}\right) \quad (9)$$

Equation (9) is the formula used in theoretical calculations. In these calculations T_h and T_c are the average bulk temperature for hot and cold sides:

$$T_h = \frac{T_{h,in} + T_{h,out}}{2} \quad (10)$$

$$T_c = \frac{T_{c,in} + T_{c,out}}{2} \quad (11)$$

2.3 Experimental study of AGMD

The MD module being tested in our laboratory was first developed by the Swedish National Development Co. ('Statsföretag', now defunct). The company decided to design a membrane distillation plastic cassette for seawater desalination systems and initiate testing under real-life conditions. The cassettes are designed so that they can be stacked together to form modules of varying size. Each cassette consists of injection molded plastic frames containing two parallel membranes, feed and exit channels for the warm water, and two condensing walls. By interconnecting the cassettes, channels for the cooling water are formed between the condensing walls of adjacent cassettes.

This module has been improved to produce ultrapure water suitable for the semiconductor industry (here purity requirements are similar to those for current high pressure boilers and other related equipment). Tests conducted at Sandia National Laboratories in 1998 [28] demonstrated that MD generally performed equally to RO in terms of water quality with a variety of feedstocks (including nonpotable city water). From March 2003 until now, Xzero AB and Department of Energy Technology, KTH have cooperated in running tests with this facility. The current facility is comprised of one membrane module which consists ten cassettes, and the total membrane area is around 2.3 m². Roughly 10000 h of operation time have been logged with the original equipment (earlier testing was performed in Karlstad). The membrane material is PTFE with a porosity of 80% and thickness of 0.2 µm. The width of air gap of AGMD is 2 mm. The membrane module is made up of nine feed channels and nine permeate channels, and the size of module is 63 cm wide and 73 cm high with a stack thickness of 17.5 cm. A photograph of the membrane module is shown in Figure 4.



Figure 4. AGMD membrane distillation module, Xzero AB (stack thickness is 35 cm in photo)

The experimental flow diagram can be seen in Figure 5. When running the experiment, two taps are used to measure the water flow rate from cold and hot sides. During the experimental period, the pilot plant was operated in three kinds of flow conditions, e.g. $1.36 \text{ m}^3/\text{h}$, $2.04 \text{ m}^3/\text{h}$, and $2.72 \text{ m}^3/\text{h}$. For simplicity identical flow rates were maintained for both hot and cold streams.

In the pilot plant the feedstock (tap water mixed with table salt) is contained in a 1000 l tank equipped with three electrical heating elements. Pumps maintain the flow rates to the desired levels. On the cold side, an expansion tank is used to accommodate the expansion of heated cooling water.

Tap water from EGI lab hall is used to cool the cold side through a heat exchanger, so the cold side water is in a closed system. In the summer time, the tap water temperature is very stable and around 25°C , and the cold side temperature can be maintained in the range of $19\text{--}25^\circ\text{C}$. In the hot and cold water pipes and water tank, the temperature of hot/cold water was recorded before entering the MD module, and data is then transferred via a PLC to a computer. The entire experimental setup can be displayed on the computer screen, and the pilot plant is operated by Citect Runtime via PLC connection/control system. With this one can start/shut down and operate the pilot plant easily (remote operation and data acquisition is also possible).

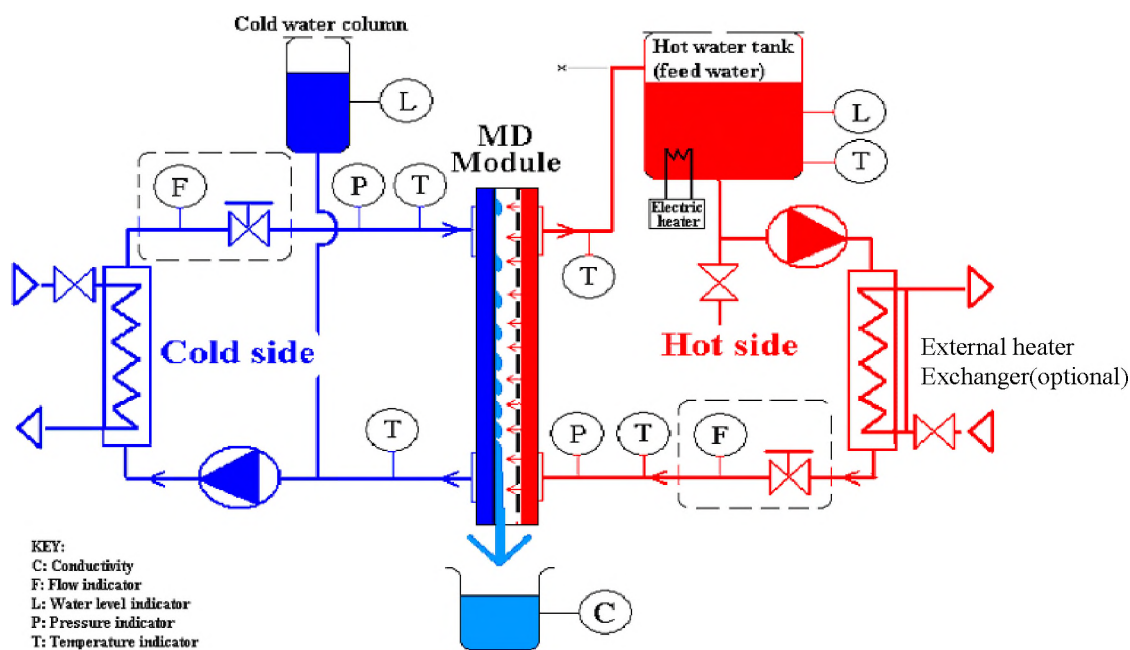


Figure 5. The experimental flow diagram in Pilot plant of Xzero at KTH

2.4 Experimental results

When conducting the experiments, there are several data that are obtained, i.e. pure water production, temperature of cold and hot sides, conductivity of pure water, TDS (total dissolved solids), pH, etc. Measurements related to water quality provide qualitative information as to the expected performance of the unit; more rigorous testing is beyond the scope of this study. TDS is measured with a TDS indicator, pH is measured with pH indicator paper, and pure water conductivity is measured with conductivity sensor. Measurement equipment is shown in Table 1. The pure water production is calculated by timing the volume of water filling a graduated cylinder.

Parameter	Measurement method/equipment
Conductivity	YOKAGAWA 402
TDS	HANNA TDS indicator
pH	pH test paper
Hardness	Hardness Test Paper

Table 1. Measurement method or equipment in Xzero Test Facility

The measurements are conducted twice per day (three measurements per data type). Recent tests have shown that the conductivity of the pure water produced by the Xzero module is around $1.5 \mu\text{S}/\text{cm}$ at steady state, regardless of feedstock salt concentration (which ranged from 300-1100 ppm). The pH value of the produced water is around 6.5-7.0. The TDS value is always under detection limits (4 mg/l). Detailed water analyses were performed in the early 1990's under similar conditions (see Appendix A for these results).

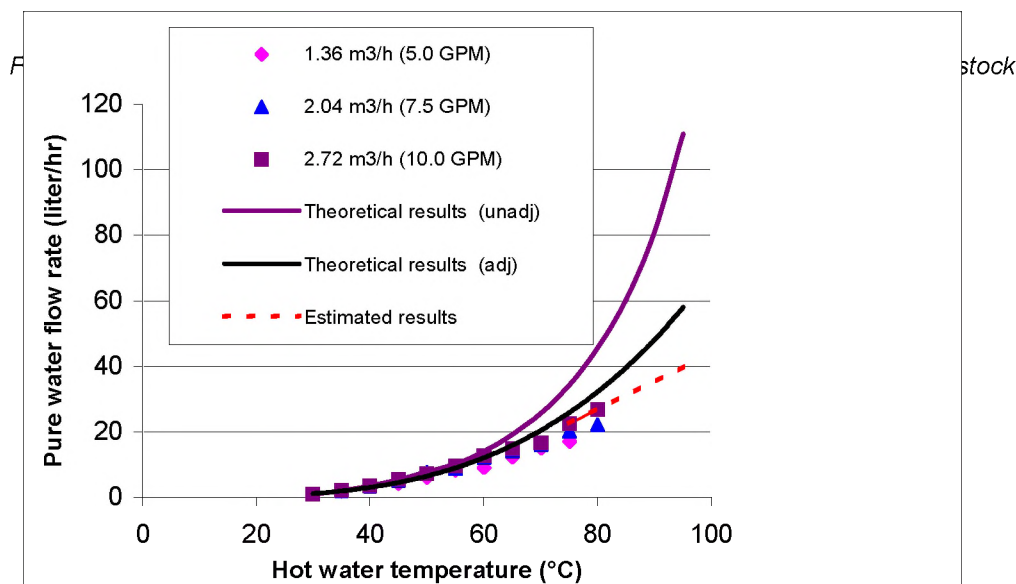


Figure 6 shows a plot of pure water production as a function of hot water temperature for the three internal recirculation flow rates tested. The performance improved at higher flow rates, when the temperature drop in the streamwise direction is lowest. Comparisons to theoretical results are also included. ‘Unadjusted’ theoretical results refer to predictions obtained from the procedure outlined in section 2.2. Significant discrepancies can be seen between this relationship and the experimental data. An attempt was made to modify theoretical predictions by accounting for heat losses and temperature polarization effects (see Appendix B for more details). This prediction, denoted ‘adjusted’ in Figure 10, appears to match experimental values at temperatures under 80°C. However performance is overpredicted at higher temperatures, even with these factors under consideration. Thus a simple linear extrapolation has been employed to predict performance beyond the range of experimental data, up to temperatures of 95°C.

3. Integrated membrane distillation with thermal cogeneration

3.1 Introduction: Parameters for case study

In order to illustrate how an MD water purification system could be employed, data from relevant cogeneration facilities has been acquired for a case study. One good example is the Örebro Cogeneration Facility, a biomass-fired unit (wood chips and peat fuels) with an installed capacity of 131 MW_e and 325 MW_{DH} [24]. A flue gas condensate unit was installed in 2002. Table 2 shows the characteristics of the condensate. The unit also includes an extensive condensate treatment system that enables the production of purified make-up water from the condensate (hence avoiding the need to purchase raw water). As shown in Figure 7, there are several steps to clean the flue gas condensate, i.e. vibration screen, UF, softening, and RO. The flue gas condensate is first treated by blowing air to drive off carbon dioxide and then is dosed with NaOH to increase pH. The permeate flow rate from the RO unit is 20 m³/h, and conductivity ranges from 2-5 µS/cm. Afterwards the permeate from RO system proceeds to the ion exchange bed to be purified to meet boiler feed water standards. The boiler make-up water demand is 5-30 m³/h and water quality requirement is shown in Table 3.

Parameter	Flue gas condensate characteristics
Temperature (°C)	57-60
Flow rate (m ³ /h)	30
pH	3.5
Organic content (%)	0.1
NH ₃ content (mg/l)	100-200

Table 2. Condensate characteristics in Örebro Power [24]

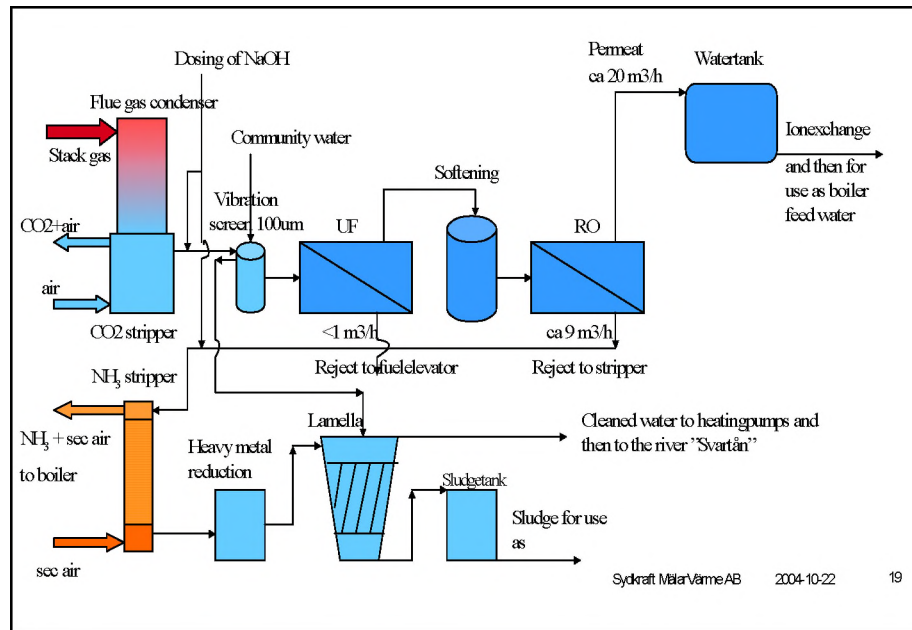


Figure 7. Flue gas condensate cleaning system [24]

Conductivity ($\mu\text{S}/\text{cm}$)	< 0.1
Silica ($\mu\text{g}/\text{l}$)	< 5
Na ($\mu\text{g}/\text{l}$)	< 5
O ₂ ($\mu\text{g}/\text{l}$)	< 5

Table 3. Feed water requirement of high pressure boiler [24]

In this system the most logical application of membrane distillation is as an alternative to the RO unit. Assuming that the water quality produced by the MD unit meets or exceeds that obtained from the RO unit, the essential issues of comparison include energy needs and economic aspects related to a system producing 10 m³/h of demineralized water (i.e. one-half the nominal capacity of the Örebro Cogeneration Facility). Predicted MD unit performance is obtained from appropriate scale-up of experimental data for a single module, as outlined in the previous section. The Fynsværket cogeneration facility in Odense is considered with regards to typical heat sources and sinks for northern European applications. Operation data is listed in Table 4. Operations vary according to the seasonal needs for district heating.

It should be noted that the coolant temperatures considered in this case study are somewhat higher than those tested in the Xzero pilot facility. This has been accounted for via the numerical model presented in the previous section. (MD performance is not very sensitive to cold-side temperatures so long as the dewpoint temperature is maintained.)

Parameter	DH water supply	DH water return
Temperature (°C)	80-96	45-50
Pressure (bar)	3.2	3.1
Mass Flow (t/h)	5800	
Make up flow (t/h)	30	
Heat supply (MW)	235-310	

Table 4: district heating network operation data at Fynsværket in Odense

3.2 MD unit connection alternatives

Aspen Utilities was employed to construct the computer calculation modules. According to operation data of the DH network, three alternatives are considered. In Alternative 1 and Alternative 2, the DH return line is used for cooling of the MD units via a heat exchanger. The DH supply line is employed as the heat source for MD. In Alternative 1, the number of MD modules is calculated with winter conditions as the baseline. Pure water production is lower during the summer since DH supply temperatures are lower (presumably this is not a problem, as such plants normally reduce overall operations in the warmer months). For Alternative 2 the size of the MD unit is determined from summer conditions, and only a portion of the MD modules are utilized during the winter in order to yield the necessary output of 10 m³/h. Alternative 3 utilizes low grade steam as heat source of MD system. Therefore a fixed, relatively high MD inlet temperature is obtained, resulting in stable water production along with a reduced membrane area, pipe size and pump power etc. Finally, all three alternatives employ heat recovery on the cooling loop, i.e. cooling is supplied by the DH return line. Thus the net thermal demand of the MD unit is determined by comparing the DH (backpressure) condenser thermal power with the baseline case where MD is not employed.

In order to minimize the pumping power requirements for the relatively high recirculation rates, the flow layout of these three alternatives is designed with a cascading concept where each stage is comprised of parallel-coupled modules. (Each module has a flowrate of 2.7 m³/h, corresponding to tested values.) For all alternatives degasified flue gas condensate is taken as the feedstock to the MD unit. Flue gas condensate is preheated by the permeate from MD units and blow down steam to minimize specific heat energy consumption. Blow-down operations have not been considered in this investigation as they are judged to have minimal impact on energy consumption and economic factors.

In these simulations it is important to properly estimate the pressure drop through the system in order to calculate pump electricity consumption. Table 5 shows the pressure drop for the system components [25, 29, 30].

System equipment	Pressure drop, bar
MD module stage	0.1
Heat exchanger	0.08
Piping	0.04
Valve and other components	0.1
Total pressure drop	0.32

Table 5: Pressure drop of system components

Alternative 1. Fixed MD modules variable pure water production

The details of this alternative are shown in Figure 8. A total of 10 cascading stages have been selected for this design, where each stage contains around 30 modules. Winter operation has been selected as the design condition for determining the number of modules required for the desired production of 10 m³/h.

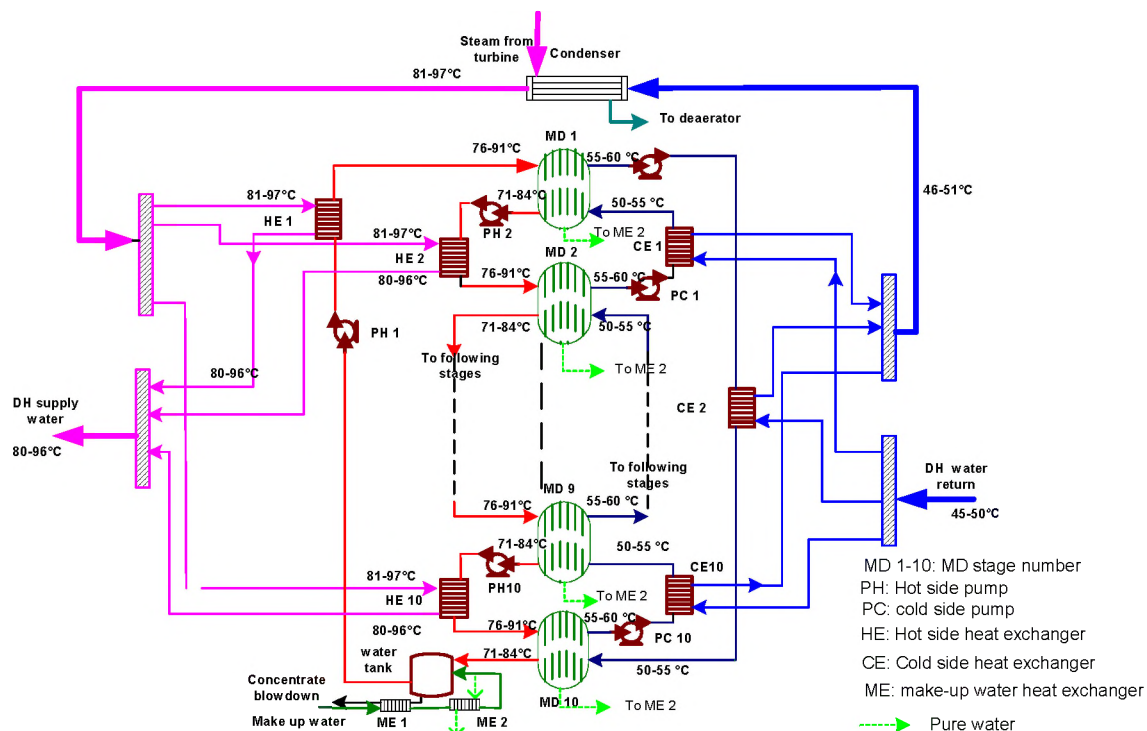


Figure 8. Connection of fixed number of MD units with DH network

Simulation results are shown in Table 6, and detailed heat/mass balance results are contained in Appendix C.

Parameter	Winter case	Summer case
DH temperature (°C)	97	81
MD hot side temperature (°C)	91	76
MD cold side temperature (°C)	55	50
Water output (m ³ /h)	10.1	5.2
Specific thermal energy consumption (kWh/m ³)	4.0	4.0
Pumping power (kW)	20	19
Specific electricity consumption (kWh/m ³)	2.0	1.9
Membrane area (m ²)	828	828

Table 6: Simulation results of Alternative 1

Alternative 2. Variable MD modules with variable pure water production

This configuration resembles Alternative 1, however the size of the MD unit is determined from summer conditions, and only a portion of the MD modules are utilized during the winter in order to yield the necessary output of 10 m³/h. Data contained in Table 7 show that the total membrane area increases by about 100 % given these conditions.

Parameter	Winter case	Summer case
DH temperature (°C)	97	81
MD hot side inlet temperature (°C)	91	76
MD cold side inlet temperature (°C)	55	50
Water output (m ³ /h)	10.1	10.1
Specific thermal energy consumption (kWh/m ³)	4.0	4.0
Pumping power (kW)	20	40
Specific electricity consumption (kWh/m ³)	2.0	4.0
Membrane area (m ²)	828	1656

Table 7: Simulation results of Alternative 2

Alternative 3. Low grade steam utilization

Since MD unit performance increases linearly with temperature, it is of interest to explore an option utilizing a higher temperature source to study the influence on membrane area. Here, low grade steam (2-3 bar, 200°C) is assumed to be available in sufficient quantities to provide heat at a level of 95°C. Figure 9 shows the layout of this alternative. The steam is supplied to an auxiliary heater, and condensate is returned to the condenser.

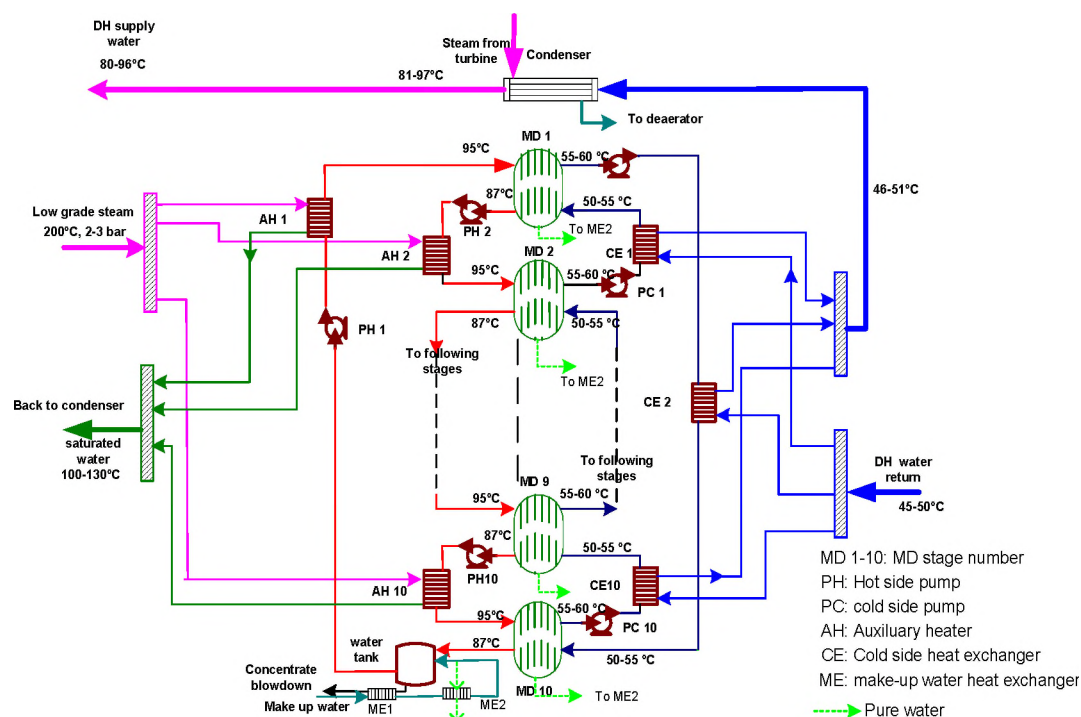


Figure 9. Connection of fixed number of MD units with an auxiliary heater

As shown in Table 8, the benefits of utilizing low grade steam include a significant reduction in membrane area as well as lowered pumping power. These gains must be weighed against the added cost and complexity of a new steam line.

Items	Winter case	Summer case
MD hot side inlet temperature (°C)	95	95
MD cold side inlet temperature (°C)	55	50
Water output (m ³ /h)	10	10
Specific thermal energy consumption (kWh/m ³)	5.0	5.0
Pumping power (kW)	15	15
Specific electricity consumption (kWh/m ³)	1.5	1.5
Membrane area (m ²)	660	660

Table 8: Simulation results with auxiliary heater (Alternative 3)

Comparison of Specific Energy Consumption

The above results show the trade-off between thermal energy and electricity as the source temperature is raised. Clearly the size of the MD unit is large as the hot-side temperature is reduced, and this is especially apparent in comparing Alternative 2 to the other options. As indicated previously, the thermal energy demand is computed from the difference in DH condenser thermal power with and without an MD unit. While heat recovery is assumed on the cooling loop, the possibility of including the permeate in a broader energy balance has not been considered. Hence the specific thermal energy consumption may actually be lower than reported. Electricity demand is linked to relatively high hot-side and cold-side recirculation rates, each roughly one order of magnitude higher than the pure water production rate. However, Alternatives 1 and 2 show specific electricity values roughly at the same level for comparable RO systems. Recirculation rates can be reduced further by a combination of optimized process layout as well as improved MD module design (this would also have an impact on thermal requirements and cost). Hence it is fair to say that the cases studied herein suggest that membrane distillation demonstrates satisfactory energy performance given the fact that the technology is not fully commercialized.

4. Cost analysis

4.1 Key information and assumptions

The following economic analysis considers the capital and operating costs directly related to the membrane distillation unit, including all necessary connections. Costs related to pretreatment equipment have not been included. The following information and assumptions have been employed [25]:

- Cogeneration facility operated 8000 h/yr.
- Demineralized water production is 10 m³/h (nominally).
- Heat sources/sinks are described in previous chapter.
- The net present value and internal rate of return is 7 %.
- Membrane lifetime is 5 yr, other equipment lifetime is 15 yr.
- Combined unforeseen expenses (i.e. margin of safety) and gross profit is 50%.
- The annual maintenance cost is 5% of the installed cost.
- Membrane price is 290 SEK/m².
- Electricity price is 0.5 SEK/kWh, and heat energy price is 0.4 SEK/kWh (taxes excluded).
- Construction price is 350 SEK/hour.
- Cost of control system for MD unit is 90 000 SEK
- The lifetime of steam line in Alternative 3 is 20 years.
- System heat losses are 5%.
- Required steam flow rate (5-10 kg/s) does not significantly impact plant operations.
- Costs related to increased space demands not included. (Alternatives 1 and 2 require about 300 m² of additional space, while Alternative 3 requires an additional 200 m².)
- Variable cost includes membrane removal and replacement at five year intervals.

4.2 Costs, Alternative 1

Table 9 contains a summary of the major costs associated with this alternative. Some comments on this information:

- ‘MD modules cost’ represents the cost to construct the MD module, excluding membranes.
- The main water pipes in hot and cold sides are DN-150.
- Control system purchased from Citect and includes necessary programming.
- ‘Others’ includes a level switch valve.
- Construction costs obtained from Carlor AB [25].

Item	Quantity	Unit Cost	Net Cost (MSEK)
Membrane	828 m ²	290 SEK/m ²	0.24
MD modules	360	5000 SEK	1.8
Heat Exchangers PH1	10	40000 SEK	0.4
Heat Exchangers PH2	10	30000 SEK	0.3
Pump	20	30000	0.6
Water tank	1	20000	0.02
Piping	500 m	600 SEK/m	0.3
Valves	100	2000	0.2
Temperature Indicators	25	2300	0.06
Pressure indicators	20	2800	0.06
Security system	4	5000	0.02
Control system			0.02
Construction	800 h	350 SEK/h	0.28
Others			0.3
Subtotal			4.6
50% margin of error/profit			2.3
Total investment			6.9

Table 9. Investment table of Alternative 1

Annual costs are summarized in Table 10.

Item	Quantity	Unit Cost	Net cost (MSEK)
Pump electricity	160 MWh	500 SEK/MWh	0.08
Heat energy	336 MWh	400 SEK/MWh	0.13
Maintenance cost			0.05
Variable cost			0.07
Total annual cost			0.33

Table 10. Annual cost of Alternative 1

4.3 Costs, Alternative 2

Table 11 contains a summary of the major costs associated with this alternative, and annual costs are listed in Table 12.

Item	Quantity	Unit Cost	Net Cost (MSEK)
Membrane	1656 m ²	290 SEK/m ²	0.48
MD modules	600	5000 SEK	3.0
Heat Exchangers PH1	10	60000 SEK	0.6
Heat Exchangers PH2	10	50000 SEK	0.5
Pump	20	40000	0.8
Water tank	1	20000	0.03
Piping	500 m	600 SEK/m	0.3
Valves	100	2000	0.2
Temperature Indicators	25	2300	0.06
Pressure indicators	20	2800	0.06
Security system	4	5000	0.02
Control system			0.02
Construction	800 h	350 SEK/h	0.28
Others			0.3
Subtotal			6.65
50% margin of error/profit			3.37
Total investment			9.98

Table 11. Investment table of Alternative 2

Item	Quantity	Unit Cost	Net cost (MSEK)
Pump electricity	200 MWh	500 SEK/MWh	0.1
Heat energy	336 MWh	400 SEK/MWh	0.13
Maintenance cost			0.06
Variable cost			0.11
Total annual cost			0.4

Table 12. Annual cost table of Alternative 2

4.4 Costs, Alternative 3

Table 13 contains a summary of the major costs associated with this alternative. Some comments on this information:

- A steam line is used to heat the hot-side feedstock up to 95 °C.
- Steam parameters: Pressure 2-3 bar, Temperature 200 °C, Flow rate: 5-10 kg/s.
- Condensed steam leads to condenser (i.e. heat is recovered).
- The steam line cost includes control system.
- Steam line price obtained from Swedpower AB[26].

Item	Quantity	Unit Cost	Net Cost (MSEK)
Membrane	660 m ²	290 SEK/m ²	0.2
MD module	287	5000	1.4
Steam line including control system	100m		1.0
Heat Exchangers PH1	10	20000 SEK	0.2
Heat Exchangers PH2	10	20000 SEK	0.2
Pump	20	12000 SEK	0.24
Water tank	1	20000	0.02
Piping	400 m	500 SEK/m	0.2
Valves	50	2000	0.1
Temperature indicators	10	2300	0.02
Pressure indicators	10	2800	0.03
Others			0.45
Security system	4	5000	0.02
Control system			0.03
Construction	600 h	350 SEK/h	0.21
Subtotal			4.32
50% margin of error/profit			2.16
Total investment			6.48

Table 13: Investment table of Alternative 3

Annual costs are summarized in Table 14.

Item	Quantity	Unit Cost	Net cost (MSEK)
Pump electricity	120 MWh	500 SEK/MWh	0.06
Heat energy	420 MWh	400 SEK/MWh	0.17
Steam line Maintenance cost			0.03
Maintenance cost			0.04
Variable cost			0.05
Total annual cost			0.35

Table 14: Annual cost table of Alternative 3

4.5 Cost calculation results

The capital costs of purified water production can be estimated based on the above data (fifteen-year operational span), and these results are summarized in Table 15. Utilizing the higher temperature heat source shows the best performance, even if the cost to install a steam line is included. At present the specific water cost for an RO system with same capacity is around 4 SEK/m³ [27]. All estimates compare fairly well to reverse osmosis systems, especially if one considers the potential for technology improvement with this young technology.

	Specific water costs, low end estimate (SEK/m ³)	Specific water costs, high end estimate including 50% uncertainty/profit (SEK/m ³)
Alternative 1	10	14
Alternative 2	14	19
Alternative 3	10	12

Table 15. Specific capital cost comparison of the three alternatives

5. Conclusions and future work

This investigation shows that membrane distillation is a promising technology for demineralized water production in thermal cogeneration plants. Experimental trials conducted to date indicate that this technology meets minimum requirements regarding proof-of-concept and water quality. Data obtained from relevant experimental trials was scaled up to illustrate the performance of a 10 m³/h unit supplied with heat either from the district heat supply line or low-grade steam. Heat and mass balance calculations show that the specific thermal and electricity demands lie around 4.0-5.0 kWh/m³ and 1.5-4.0 kWh/m³, respectively. Although the combined energy consumption is higher than reverse osmosis, the ability to recover heat coupled to expected improvements in process and component design make MD a competitive alternative. Economic estimates place the cost of demineralized water production in the range of 10-14 SEK/m³ for the most likely scenarios; this level lies above RO but is nonetheless positive given the early stage of development of MD technology.

These results suggest that follow-on studies are of merit, especially for applications where the raw water quality is much lower than that studied in the present investigation. Desalination is clearly one promising area that deserves more attention. For cogeneration applications, the construction and testing of a pilot plant with a capacity of at least 1 m³/h is suggested. Pilot plant trials will allow for a more detailed evaluation of this technology under actual operating conditions (especially at higher temperature levels which could not be tested in the present experimental facility). Specific aspects related to water quality issues and long-term performance, which were largely beyond the scope of the present investigation, can be readily addressed in these tests. Options for commercialization, including more detailed economic analyses, should also be considered. Finally there is a clear need to conduct in-depth fundamental heat and mass transfer studies on actual membrane modules in order to further optimize the process.

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Appendices

A Water quality analysis report from KeMiska Stationen AB, Dated April 4, 1991

JÄMFÖRELSE AV ANALYSRESULTAT

PARAMETER	KRANVATTEN	RENAT VATTEN	TJÄNLIGT FRÅN VATTENVERK
FÄRG	5	<5	<15
TURBIDITET	0.4	0.2	<0.5
LUKT	INGEN	TYDLIG	INGEN
BOTTENSATS	INGEN	INGEN	INGEN
KONDUKTIVITET	16.2	0.6	SAKNAS VÄRDE
COD-Mn	2	<1	<4
pH	7.8	7.2	7.5 - 9.0
AMMONIUM-N	<0.1	<0.1	<0.4
TOTALHÅRDHET	3.9	0	<15
KALCIUM	25	0	20 - 60
MAGNESIUM	1.9	0	<30
NATRIUM	9.2	0.07	<100
KALIUM	2.3	0.5	<20
JÄRN	<0.05	<0.05	<0.10
MANGAN	<0.02	<0.02	<0.05
ALKALINITET	50	1	>60
KLORID	11	0.3	<100
FLUORID	0.2	<0.1	0.8 - 1.2
SULFAT	39	<1	<100
NITRAT-N	<0.5	<0.5	<5
NITRIT-N	0.003	0.002	<0.005
FOSFAT-P	<0.05	<0.05	<0.20
ALUMINIUM	0.04	0.03	<0.10
KOPPAR	<0.02	<0.02	<0.05

JÄMFÖRELSE AV ANALYSRESULTAT

PARAMETER	SALTVATTEN RÅVATTEN	RENAT VATTEN	TJÄNLIGT FRÅN VATTENVERK
FÄRG	15	5	<15
TURBIDITET	1.4	0.7	<0.5
LUKT	INGEN	BRÄNT	INGEN
BOTTENSATS	SPÅR	INGEN	INGEN
KONDUKTIVITET	281	7.8	SAKNAS VÄRDE
COD-Mn	6	(74)	<4
pH	7.8	4.6	7.5 - 9.0
AMMONIUM-N	<0.1	<0.1	<0.4
TOTALHÅRDHET	22.4	0.4	<15
KALCIUM	44	1.9	20 - 60
MAGNESIUM	70	0.7	<30
NATRIUM	500	2.8	<100
KALIUM	27	0.3	<20
JÄRN	<0.05	<0.05	<0.10
MANGAN	<0.02	<0.02	<0.05
ALKALINITET	69	0	>60
KLORID	970	9	<100
FLUORID	0.3	<0.1	0.8 - 1.2
SULFAT	165	<1	<100
NITRAT-N	0.6	<0.5	<5
NITRIT-N	0.006	0.005	<0.005
POSFAT-P	<0.05	<0.05	<0.20
ALUMINIUM	0.12	<0.02	<0.10
KOPPAR	<0.02	0.03	<0.05

JÄMFÖRELSE AV ANALYSRESULTAT

PARAMETER	SÖTVATTEN RÅVATTEN	RENAT VATTEN	TJÄNLIGT FRÅN VATTENVERK
FÄRG	30	10	<15
TURBIDITET	13	0.5	<0.5
LUKT	SVAG OBEST	BRÄNT	INGEN
BOTTENSATS	TÄML STOR	INGEN	INGEN
KONDUKTIVITET	34.6	4.5	SAKNAS VÄRDE
COD-Mn	12	(12)	<4
pH	8.5	7.7	7.5 - 9.0
AMMONIUM-N	<0.1	<0.1	<0.4
TOTALHÅRDHET	6.9	0.4	<15
KALCIUM	38	2.0	20 - 60
MAGNESIUM	7.0	0.4	<30
NATRIUM	34	4.6	<100
KALIUM	4.3	0.5	<20
JÄRN	0.22	<0.05	<0.10
MANGAN	0.08	0.03	<0.05
ALKALINITET	111	8	>60
KLORID	62	9	<100
FLUORID	0.3	<0.1	0.8 - 1.2
SULFAT	41	<1	<100
NITRAT-N	<0.5	<0.5	<5
NITRIT-N	0.010	<0.002	<0.005
FOSEFAT-P	<0.05	<0.05	<0.20
ALUMINIUM	0.09	<0.02	<0.10
KOPPAR	<0.02	<0.02	<0.05

B Non-ideal behavior of MD

There are three major heat and mass transfer inefficiencies in MD, i.e. temperature polarization across the membrane, mass transfer resistance to vapor flow through the membrane due to the presence of trapped air in the pores, and conductive heat losses through the membrane [19]. More information on these losses are provided below, and other details can be found in Chuangfeng [31]:

B.1 Temperature polarization effects

Heat is transferred across the membrane through two primary pathways: latent heat of evaporation, together with convective transport in vapor phase, and conduction heat transfer through the membrane. Thermal boundary layers develop as a result of these resistances, ultimately limiting mass transfer [32]. One way to quantify these effects is via the so-called temperature polarization coefficient, *TPC* [32]:

$$TPC = \frac{T_{m1} - T_{m2}}{T_{b1} - T_{b2}} \quad (1)$$

where T_{m1} is the interfacial feed temperature, T_{m2} the interfacial permeate temperature, T_{b1} the bulk feed temperature, and T_{b2} is the bulk permeate temperature. This value represents the ratio of the actual driving force to the overall driving force. A schematic drawing of temperature polarization in MD is shown in Figure A1. Since the evaporation and condensation rates depend on the interfacial temperatures (not the bulk temperatures), and because the partial vapor pressure driving force is mainly a function of temperature, it is desired that the difference between T_{m1} and T_{m2} should be as high as possible [33]. In other words, the *TPC* should be as close to unity as possible. However, in practice, the *TPC* value is lower than unity and varies between 0.2 and 0.9[34], depending on the membrane module configuration. Results from Phattaranawik et al. [35] show that a *TPC* of about 0.4 is suitable for the present case (PTFE membrane, thickness of 0.2 μm). Higher coefficients are possible via enhanced turbulence [36], although tradeoffs with pressure drop and reduced surface areas must be properly accounted for.

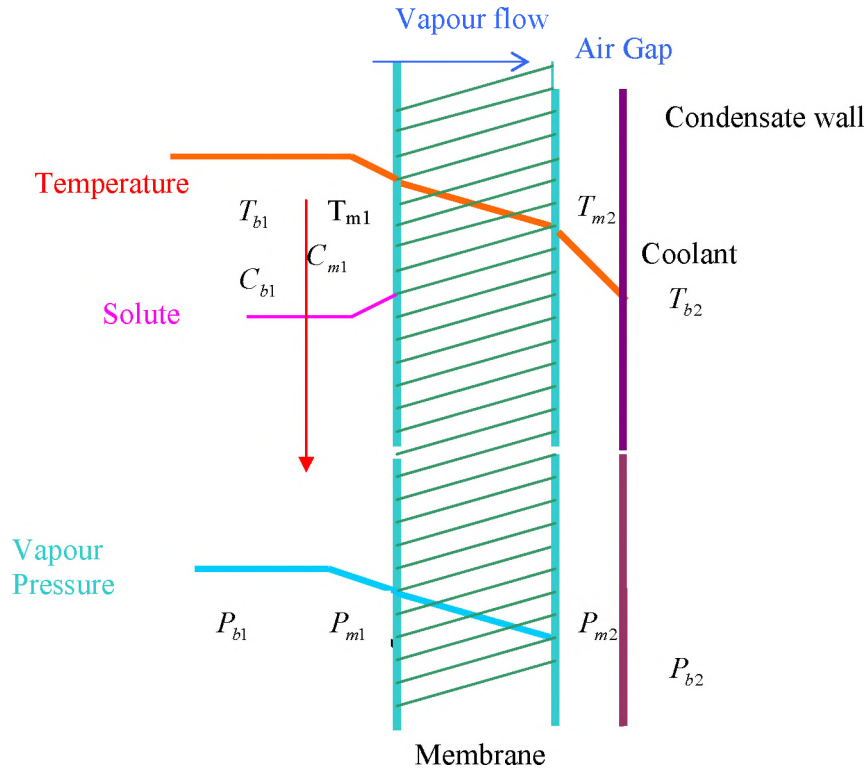


Figure A1. Profile diagram of air gap membrane distillation

B.2 Resistance to mass transfer through the membrane in MD

Vapor transported through membrane pores encounter resistance from trapped air/gases and via the tortuosity of the pore structure [38]. Several approaches to overcome these resistances have been studied in the last twenty years. In the early 1980s, Schneider and van Gassel (1984) introduced the idea of degasification of the working fluids in DCMD systems to reduce the partial pressure of air in the membrane pores. Membrane properties such as porosity, pore size, tortuosity, and membrane thickness are most important factors to dictate the resistance to mass transfer through micropores in MD [36]. Furthermore, studies have shown that the relationships between the membrane properties (not only the properties by themselves) as well as membrane conditions (e.g., compaction, wetting, etc.) are important for efficient operation. Therefore, selection of an appropriate membrane that poses the least resistance to mass transfer is crucial.

B.3 Conductive heat loss through the membrane

It is impossible to completely control the conductive heat loss through the membrane because of the tradeoff between a thick membrane for better heat insulation and a thin membrane for reduced mass flow resistance in the membrane [34]. One possible solution is to increase the porosity of the membrane, since the air in the pores of membrane has better insulation properties than the polymeric material of the membrane. (Increased porosity also increases the available surface area for evaporation.) However, microporous membranes with porosities of 80% and even higher already exist, and further improvement may only be achieved when nanotubetechnology matures [39]. Alternative materials for membrane are also not expected to improve the process because of the strict requirement for hydrophobicity and relatively similar heat conductivities of most hydrophobic polymers [40].

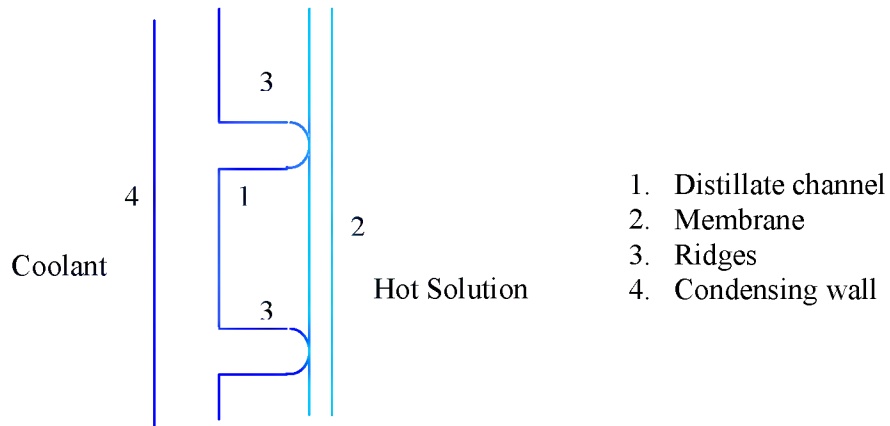


Figure A2. The ridge arrangement

Module construction also plays an important role in conductive losses. As illustrated in Figure A2, ridges used to support the membrane offer a pathway for heat transfer and block a portion of the membrane surface area for mass transfer. The latter was accounted for by an effective area coefficient, A_{eff} , defined as the unblocked membrane area divided by the actual membrane area. A value of $A_{eff} = 0.9$ was assumed for the present study.

C Heat balance at site condition

C.1 Alternative 1

		Winter case	Summer case
District heating water			
DH water inlet temperature	°C	50	45
DH water mass flow	m ³ /h	5800	5800
DH water inlet pressure	bar	3.2	3.2
DH water temperature after CE (1-10)	°C	51	46
DH water temperature before condenser	°C	51	46
DH water temperature after condenser	°C	97	81
DH water temperature after HE (1-10)	°C	96	80
DH water pressure	bar	3.3	3.3
MD unit characteristics			
Hot side water inlet temperature per stage	°C	91.0	76.0
Hot side water outlet temperature per stage	°C	84.0	71.0
Hot side water pressure	bar	1.34	1.34
Cold side water inlet temperature per stage	°C	55	50
Cold side water outlet temperature per stage	°C	60	55
MD process water flow rate	kg/s	22	22
Hot side heat demand per stage	kW	656	369
Cold side heat recovered per stage	kW	644	366
Permeate flow rate	m ³ /h	10.1	5.1
Net thermal power delivered by DH condenser*	kW	40	20
Pump power	kW	20	19
Membrane area	m ²	880	880

*as compared to baseline case without MD unit

C.2 Alternative 2

		Winter case	Summer case
District heating water			
DH water inlet temperature	°C	50	45
DH water mass flow	m ³ /h	5800	5800
DH water inlet pressure	bar	3.2	3.2
DH water temperature after CE (1-10)	°C	51	46
DH water temperature before condenser	°C	51	46
DH water temperature after condenser	°C	97	80.5
DH water temperature after HE (1-10)	°C	96	80
DH water pressure	bar	3.3	3.3
MD unit characteristics			
Hot side water inlet temperature per stage	°C	91.0	76.0
Hot side water outlet temperature per stage	°C	84.0	71.0
Hot side water pressure	bar	1.34	1.34
Cold side water inlet temperature per stage	°C	55	50
Cold side water outlet temperature per stage	°C	60	55
MD process water flow rate	kg/s	22	44
Hot side heat demand per stage	kW	656	643
Cold side heat recovered per stage	kW	644	640
Permeate flow rate	m ³ /h	10.1	10.1
Net thermal power delivered by DH condenser*	kW	40	40
Pump power	kW	20	40
Membrane area	m ²	880	1656

*as compared to baseline case without MD unit

C.3 Alternative 3

		Winter time	Summer time
District heating water			
DH water inlet temperature	°C	50	45
DH water mass flow	m ³ /h	5800	5800
DH water inlet pressure	bar	3.2	3.2
DH water temperature after CE (1-10)	°C	51	21
DH water temperature before condenser	°C	51	21
DH water temperature after condenser	°C	96	81
DH water pressure	bar	3.3	3.3
MD unit characteristics			
Hot side water temperature inlet	°C	95	95
Hot side water temperature outlet	°C	86	86
Hot side water pressure	bar	1.34	1.34
Cold side water inlet temperature per stage	°C	55	50
Cold side water outlet temperature per stage	°C	60	55
MD process water flow rate	kg/s	18	18
Hot side heat demand per stage	kW	841	836
Cold side heat recovered per stage	kW	828	830
Permeate flow rate	m ³ /h	10	10
Net thermal power delivered by DH condenser*	kW	50	50
Pump power	kW	15	15
Membrane area	m ²	660	660

*as compared to baseline case without MD unit

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The undersigned hereby certifies that the report 'Membrane Distillation and Applications for Water Purification in Thermal Cogeneration - A Prestudy' meets generally accepted quality guidelines for engineering research. These guidelines have been checked within the framework of an internal review process conducted by Tekn. Lic. Per Almqvist and the undersigned.

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Director, KTH Energy Center

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